1981

Five heat integration configurations for binary distillation systems

Teh-ping Chiang
Lehigh University

FIVE HEAT INTEGRATION CONFIGURATIONS FOR BINARY DISTILLATION SYSTEMS

by

Teh-ping Chiang

A Research Report
Presented to the Graduate Faculty
of Lehigh University
in Partial Fulfillment
of Requirements for
the Degree of
Master of Science
in
Chemical Engineering

Lehigh University
1981

ACKNOWLEDGEMENTS

I dedicated this paper to my parents.

My special thanks are due to my advisor, Professor W. L. Luyben, for his encouragement and patient guidance.

I am grateful to the Chemical Engineering Department for providing the financial support for this work.

This research report is accepted and approved in partial fulfillment of the requirements for the degree of Master of Science.

12/23/81
Date

William L Lupfer
Professor in Charge

Chairman of Department

ABSTRACT

Five designs for heat-integrated distillation systems were studied for a binary separation. The five different heat-integration configurations were compared with a conventional single column system. Energy comsumption were compared assuming the same total number of trays in all of the heat-integrated systems, the same product purities, and same capacity.

The specific chemical system studied was the methanol-water separation with high purities. Feed compositions from 30 to 80 mole% methanol were explored.

Three of the schemes were found to give quite similar advantages over the single conventional column. Configuration 4 was found to be the best energy saving scheme , especially for low feed compositions. Energy saving ranged from 30 to 50 % .

INTRODUCTION

Energy prices have increased drastically in the last ten years. Since distillation processes are major energy consumers in many plants it has become desirable to make those operations use energy more efficiently. Such improvement will often require additional capital investment.

Many energy conservation schemes have been known for some time. They were reviewed by Robinson and Gilliland (1950) and more recently by King (1971) . These schemes have not been used frequently in the past because their higher capital investment could seldom be justified when energy costs were low.

Freshwater (1951) classified these various schemes under three catagories : multiple effects or heat integration methods, vapor recompression or heat pump methods and indirect methods. Most methods fall under the first two catagories. Rathore (1974) discussed synthesis strategies for multicomponent separation with energy integration. Tyreus (1975) explored the control of one heat-integration system. Morari (1980) has studied the dynamic aspects of several heat-integrated schemes.

Johnson (1971) has studied the modeling of vapor recompression separation using modular approach. Robertson (1974) recommended this method for propylene splitter.

Null (1976) presented guidelines for conditions under which vapor recompression might be an economical substitute for conventional distillation process design.

In this report, the steady state designs of five heat-integrated configurations were compared with a single conventional column on the basis of energy consumption.

A steady state computer program was written which designed a two-column heat-integrated system for a given feed and specified product compositions. Rigorous plate-to-plate calculations were made for those systems taking into account the effects of non-equal molar overflow and nonideal vapor-liquid equilibrium. The assumptions made for this study are :

(1) No pressure drop in the column.

(2) Ideal gas.

(3) Heat loss amounts to 5 % the total heat input.

(4) Tray efficiency is 75 %.

(5) Binary methanol/water system.

(6) Product purities X_D = 99.9 mole%

X_B = 0.1 mole%

CONFIGURATIONS STUDIED

(I) Base Case

A conventional single column system, see Figure 1 , was the base case in this report. Energy is supplied to the reboiler and removed from the condenser. Because of the temperature difference between the reboiler and the condenser, the separation of components is always accompanied by a degradation of energy, even when heat leaks and other losses are excluded. Estimates as low as 1.9% have been reported for the thermodynamic efficiency of industrial distillation columns (Freshwater , 1951).

(II) Heat-Integration Systems

The basic feature of a heat-integrated system is to utilize the heat content of the overhead vapor generated in one column to supply the heat required in the reboiler of the other column. In order to provide the necessary temperature difference the columns must be operated under different pressures. Three limitations should be noted. First, temperatures and pressures can not approach too closely to the critical temperature or pressure. Second, temperature-sensitive materials may limit maximum base temperatures. Third, overhead vapor temperature from high pressure column should operate at a pressure which will generate a reasonable temperature difference from the reboiler temperature of low pressure column so that heat transfer areas are not excessive. In the methanol-water system the high pressure column was operated at 100 psia. The low pressure column was operated at 17 psia. This gave a temperature difference for heat transfer of 36 °F.

A. Configuration 1

Feed is split more-or-less equally between the two columns. The overhead vapor product of high pressure column is used to supply the heat required in the low pressure column, see Figure 2. This is the configuration studied by Tyreus.

In the following, this configuration was abbreviated FS, i.e. Feed-Split .

B. Configuration 2

All the feed is fed to the high pressure column. About half of the methanol product is removed overhead from the high pressure column. The bottoms product from the high pressure column is fed into the low pressure column. Distillate product composition in both columns and bottoms product composition of low pressure column are at the desired purities. Heat-integration is in the same direction as process flows. We therefore call this heat integration in the forward direction, see Figure 3.

In the following, this configuration was abbreviated LS/F, i.e. Light-Split-Heat-Integration-Forward.

C. Configuration 3

The difference between configuration 3 and configuration 2 is that the distillate product from high pressure column is fed into the low pressure column and about half of the final water product is removed from the bottoms of the high pressure column. Bottoms product compositions of both columns and the distillate product composition of the low pressure column are at the desired purities. Heat-integration is also forward , see Figure 4 .

This configuration was abbreviated HS/F , i.e. Heavy-Split-Heat-Integration-Forward.

D. Configuration 4

All the feed is fed into the low pressure column. Bottoms product from low pressure column is fed into high pressure column. Heat integration is from the down-stream high pressure column backward to the up-stream low pressure column, see Figure 5 . In this case heat integration is in the direction opposite to the direction of the process flows. Because the base of the low pressure column contains a lot of methanol , its base temperature is low. Therefore the pressure in the high pressure column could be lower and still have the same heat transfer temperature difference. 80 psia was used in this study.

This configuration was abbreviated LS/R , i.e. Light-Split-Heat-Integration-Reverse.

E. Configuration 5

The high pressure column in this case is operated as a prefractionator. Both distillate products and bottoms products, which are only partially purified, are fed into the low pressure column at different feed trays. Specification products are produced in the low pressure column. Heat-integration is in the forward direction, see Figure 6.

This configuration was abbreviated PF/F .

DESIGN METHODS

(I) Single Conventional Column

Column was designed as follows.

The minimum reflux ratio was found by extending the VLE tie line through the feed point on an enthalpy-composition diagram to the intersection with a vertical line through the distillate composition. Feed could be subcooled liquid, a vapor-liquid mixture, or superheated vapor. The actual reflux ratio was set a 1.2 times the minimum. The number of trays in the column was calculated by plate-to-plate calculations from the specified bottoms composition to the specified distillate purity. See Appendix B for detailed computational methods.

(II) Heat-Integration Systems

A. FS

Columns were designed individually using the same methods as used for a single column described above. The split ratio between the two columns was changed until the heat removal rate in the condenser of the high pressure column equaled the heat addition rate in the reboiler of the low pressure column. The total tray numbers of the two columns were used as the designed tray number for all other heat-integration systems.

B. LS/F

The amount of the light component removed in the first column was changed until the heat removal rate in high pressure column was equal to the heat addition rate in the low pressure column.

C. HS/F

The amount of the heavy component removed in the first column was changed until heat duties in the two columns were balanced.

D. LS/R

Same as LS/F .

E. PF/F

Because both distillate product and bottoms product in the high pressure column were fed into the low pressure column , this system is a dual feeds design problem.

First, the distillate product composition of high pressure column was assumed. Distillate flow rate from the prefractionator column was changed until the heat removal rate in the condenser of the high pressure column equaled to the heat addition rate in the reboiler of the low pressure column. Then prefractionator distillate composition was varied until the optimum value was found, i.e. the point where the energy consumption was minimized.

RESULTS & DISCUSSION

Feed compositions ranging from 30 to 80 mole% were explored for each case. Table 1 shows the energy requirements , Q_R , and reboiler temperature , T_R , for each configuration. Detailed steady state specifications are listed in Table 2 - 7 .

Configuration 1, 2 and 4 show fairly similar advantages over a conventional single column. Configuration 4 (LS/R) , the reverse heat-integrated scheme, gives the lowest energy consumption. Note, however , that the reboiler temperature for configuration 4 are somewhat higher than Configuration 2, requiring higher temperature heat input medium. Thus Configuration 2 might be prefered if the lowest possible pressure steam is to be used. This temperature difference is the largest at high methanol feed compositions.

The energy requirements of PF/F are for a distillate composition in high pressure column of 0.98 . When this value is increased from 0.98 up to 0.997 this system reduces to the LS/F configuration. Energy requirements as shown in Figure 7 - 9 for different distillate compositions as feed composition ranged from 30 to 80% . The limiting case of PF/F will be LS/F .

Table 1

Energy Requirements for Each Case

Z_F(mole fr.)			0.8	0.5	0.3
	Base Case	Q_R(10^6Btu/hr)	54.7	36.7	25.7
		T_R(°F)	219	219	219
1	FS	Q_R(10^6Btu/hr)	34.6	23.9	18.9
		T_R(°F)	327	327	327
2	LS/F	Q_R(10^6Btu/hr)	34.8	24.5	20.6
		T_R(°F)	264	280	295
3	HS/F	Q_R(10^6Btu/hr)	55.5	38.1	26.6
		T_R(°F)	327	327	327
4	LS/R	Q_R(10^6Btu/hr)	33.9	20.3	15.2
		T_R(°F)	311	311	311
5	PF/F	Q_R(10^6Btu/hr)	52.6	35.1	25.5
		T_R(°F)	275	293	306

For lower purity separation, X_D=99% ,FS,LS/F and LS/R still shows advantages using heat integration method. Table 8 showed the energy requirements for each configuration. The steady state specifications of LS/R , which still is the best, were listed in Table 9. PF/F was excluded.

Table 2 - Steady State Design Specifications (Base Case)

Feed composition (mole fr.) Item	0.8	0.5	0.3
Feed rate (moles/hr)	2300	2300	2300
Distillate rate (moles/hr)	1841*	1150	689
Distillate composition	0.999	0.999	0.999
Bottoms rate (moles/hr)	459	1150	1611
Bottoms composition	0.001	0.001	0.001
Operating pressure (psia)	17	17	17
Number of trays	75	99	64
Feed tray location	12	16	18
Reflux ratio	0.85	0.93	1.15
Reboiler heat duty (10^6 Btu/hr)	54.7	36.7	25.7
Reboiler temperature (°F)	219	219	219
Steam pressure (psig)	20	20	20
Reflux drum temperature (°F)	155	155	155

* = 500 million lb. methanol/year

Table 3 - Steady State Design Specifications (FS)

Feed composition (mole fr.)	0.8		0.5		0.3	
Item	Column1	Column2	Column1	Column2	Column1	Column2
Feed rate (moles/hr)	1027	1273	1097	1203	1204	1096
Distillate rate (moles/hr)	822	1019	548.5	601.5	361	328
Distillate composition	0.999	0.999	0.999	0.999	0.999	0.999
Bottoms rate (moles/hr)	205	254	548.5	601.5	843	768
Bottoms composition	0.001	0.001	0.001	0.001	0.001	0.001
Operating Pressure (psia)	100	17	100	17	100	17
Number of trays	75	45	99	42	64	28
Feed tray location	12	10	10	9	11	8
Reflux ratio	1.5	0.96	1.46	1.11	1.72	1.73
Reboiler heat duty (10^6Btu/hr)	34.6	32	23.9	20.9	18.9	15.2
Reboiler temperature (°F)	327	219	327	219	327	219
Steam pressure (psig)	150	---	150	---	150	---
Reflux drum temperature (°F)	255	155	255	155	255	155

Table 4 - Steady State Design Specifications (LS/F)

Feed composition (mole fr.)	0.8		0.5		0.3	
Item	Column1	Column2	Column1	Column2	Column1	Column2
Feed rate (moles/hr)	2300	1473	2300	1764	2300	1930
Distillate rate (moles/hr)	827	1014	536	614	370	319
Distillate composition	0.999	0.999	0.999	0.999	0.999	0.999
Bottoms rate (moles/hr)	1473	459	1764	1150	1930	1611
Bottoms composition	0.688	0.001	0.35	0.001	0.16	0.001
Operating pressure (psia)	100	17	100	17	100	17
Number of trays	75	45	99	42	64	28
Feed tray location	4	9	3	10	4	10
Reflux ratio	1.47	1.02	1.43	1.23	1.64	2.45
Reboiler heat duty (10^6Btu/hr)	34.8	31.7	24.5	20.3	20.6	15.1
Reboiler temperature (°F)	264	219	280	219	295	219
Steam pressure (psig)	55	---	70	---	90	---
Reflux drum temperature (°F)	255	155	255	155	255	155

Table 5 - Steady State Design Specifications (HS/F)

Feed composition (mole fr.)	0.8		0.5		0.3	
Item	Column1	Column2	Column1	Column2	Column1	Column2
Feed rate (moles/hr)	2300	1929	2300	1340	2300	995
Distillate rate (moles/hr)	1929	1841	1340	1150	995	689
Distillate composition	0.95	0.999	0.86	0.999	0.69	0.999
Bottoms rate (moles/hr)	371	88	960	190	1305	306
Bottoms composition	0.001	0.001	0.001	0.001	0.001	0.001
Operating pressure (psia)	100	17	100	17	100	17
Number of trays	45	75	42	99	28	64
Feed tray location	21	17	24	14	23	11
Reflux ratio	0.68	0.77	0.49	0.81	0.2	0.91
Reboiler heat duty (10^6Btu/hr)	55.5	51.0	38.1	32.6	26.6	20.4
Reboiler temperature (°F)	327	219	327	219	327	219
Steam pressure (psig)	150	---	150	---	150	---
Reflux drum temperature (°F)	256	155	262	155	270	155

Table 6 - Steady State Specifications (LS/R)

Feed composition (mole fr.)	0.8		0.5		0.3	
Item	Column1	Column2	Column1	Column2	Column1	Column2
Feed rate (moles/hr)	2300	1287	2300	1699	2300	1945
Distillate rate (moles/hr)	1012	828	601	549	355	334
Distillate composition	0.999	0.999	0.999	0.999	0.999	0.999
Bottoms rate (moles/hr)	1287	459	1699	1150	1945	1611
Bottoms composition	0.64	0.001	0.32	0.001	0.17	0.001
Operating pressure (psia)	17	80	17	80	17	80
Number of trays	45	75	42	99	28	64
Feed tray location	4	10	3	15	3	21
Reflux ratio	0.93	1.42	1.05	1.39	1.44	1.93
Reboiler heat duty (10^6Btu/nr)	33.9	31.0	20.3	24.1	15.2	19.4
Reboiler temperature (°F)	166	311	178	311	189	311
Steam pressure (psig)	---	120	---	120	---	120
Reflux drum temperature (°F)	155	240	155	240	155	240

Table 7 - Steady State Specifications (PF/F)

Feed composition (mole fr.)	0.8		0.5		0.3	
Item	Column1	Column2	Column1	Column2	Column1	Column2
Feed rate (moles/hr)	2300	756,1544	2300	1389,911	2300	1765,535
Distillate rate (moles/hr)	1544	1841	911	1150	535	689
Distillate composition	0.98	0.999	0.98	0.999	0.98	0.999
Bottoms rate (moles/hr)	756	459	1389	1150	1765	1611
Bottoms composition	0.43	0.001	0.18	0.001	0.09	0.001
Operating pressure (psia)	100	17	100	17	100	17
Number of trays	45	75	99	42	64	28
Feed tray location	9	5,23	12	5,20	16	6,18
Reflux ratio	1.01	0.7	1.11	0.75	1.35	1.04
Reboiler heat duty (10^6Btu/hr)	52.6	48.6	35.1	30.1	25.5	19.6
Reboiler temperature (°F)	275	219	293	219	306	219
Steam pressure (psig)	65	---	85	---	110	---
Reflux drum temperature (°F)	255	155	255	155	255	155

Table 8 - Energy Requirements for Lower Purity Columns

(X_D=0.99, X_B=0.001, Z_F=0.8)

System	Base Case	FS	LS/F	HS/F	LS/R
$Q_R(10^6$Btu/hr)	52.5	32.8	33.4	48.7	32.5

Table 9 - Steady State Specifications (LS/R)

(Lower Purity Separation)

Item	Column1	Column2
Feed rate (moles/hr)	2300	1258
Feed ~~composition~~	0.8	0.64
Distillate rate (moles/hr)	1042	816
Distillate composition	0.99	0.99
Bottoms rate (moles/hr)	1258	442
Bottoms composition	0.64	0.001
Operating pressure (psia)	17	80
Number of trays	25	36
Feed tray location	5	10
Reflux ratio	0.78	1.34
Reboiler heat duty (10^6Btu/hr)	32.5	29.6
Reboiler temperature (°F)	166	311
Steam pressure (psig)	---	120
Reflux drum temperature (°F)	156	240

Conclusions :

LS/R is the best energy saving scheme both for high feed composition as well as for low feed composition.

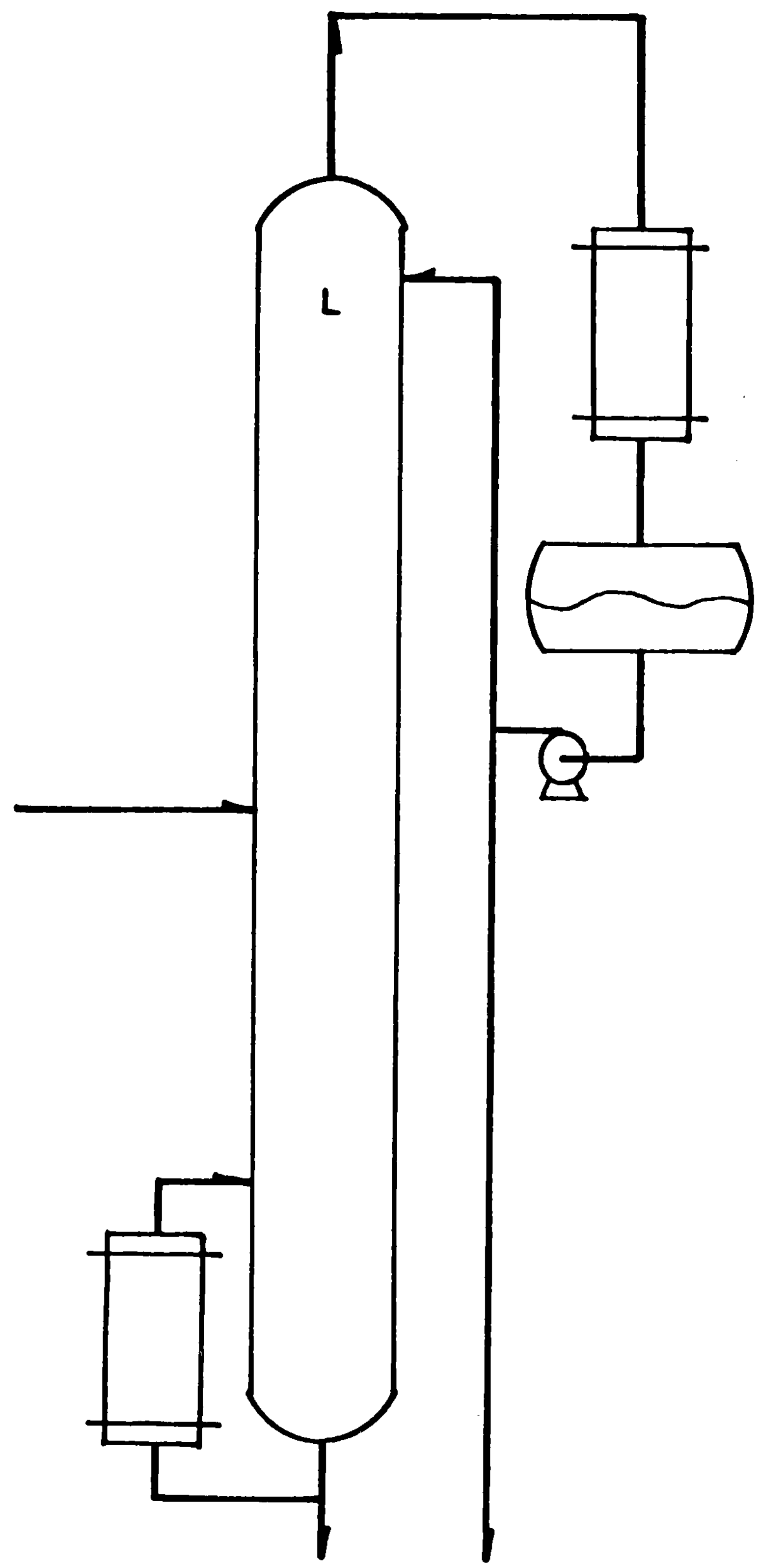

Figure - I Base Case - Conventional Single Column

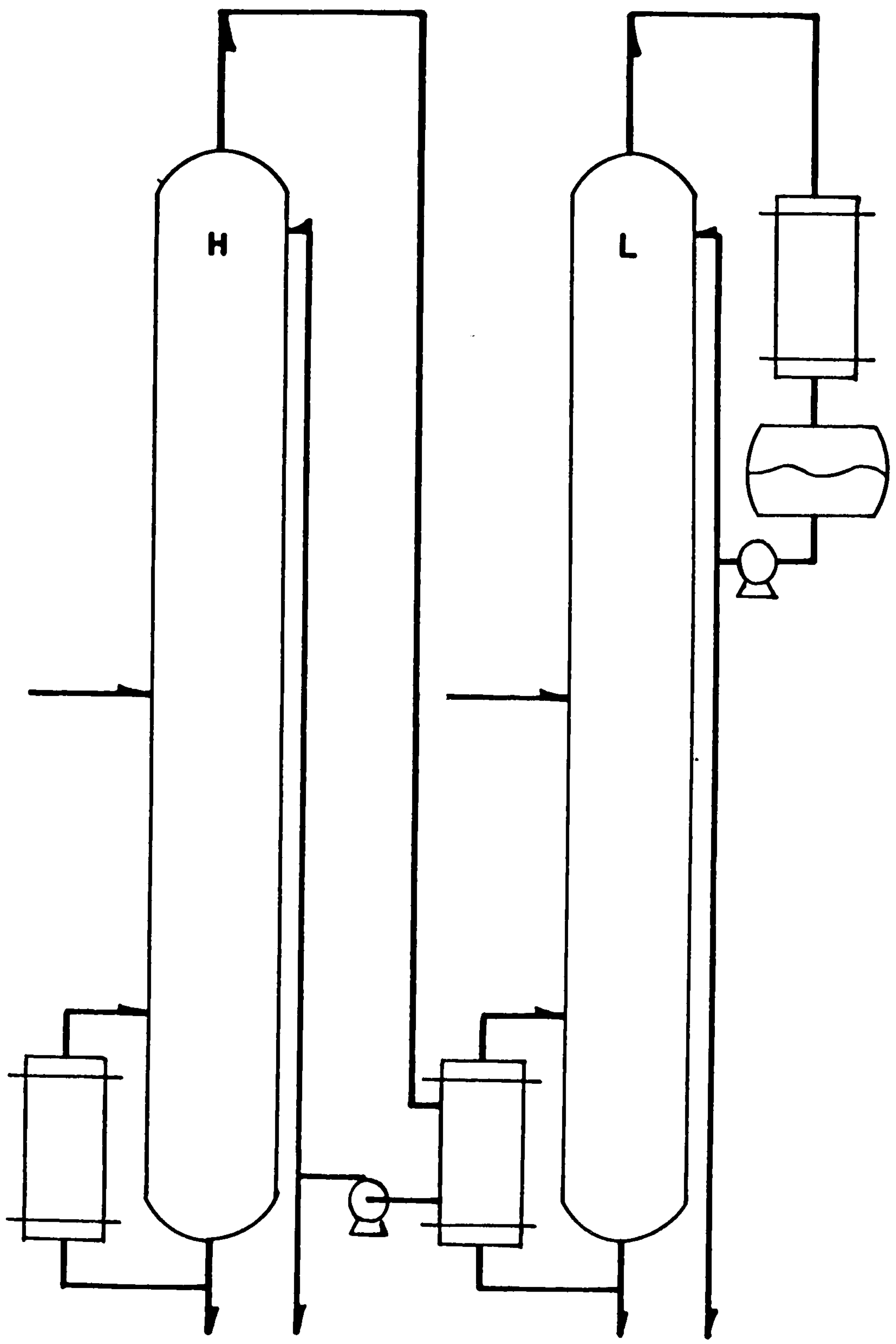

Figure - 2 Configuration I (FS)

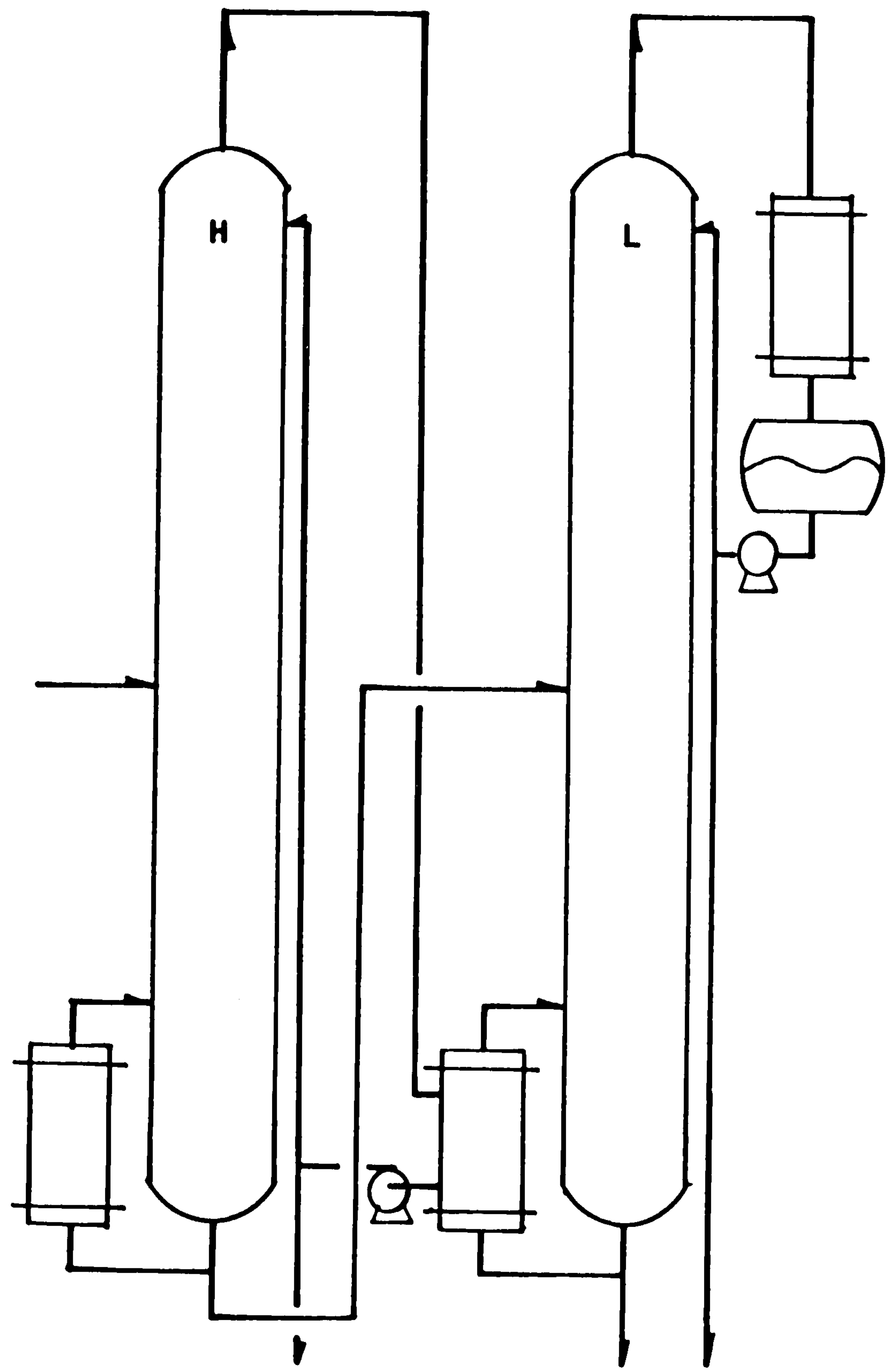

Figure - 3 Configuration 2 (L S / F)

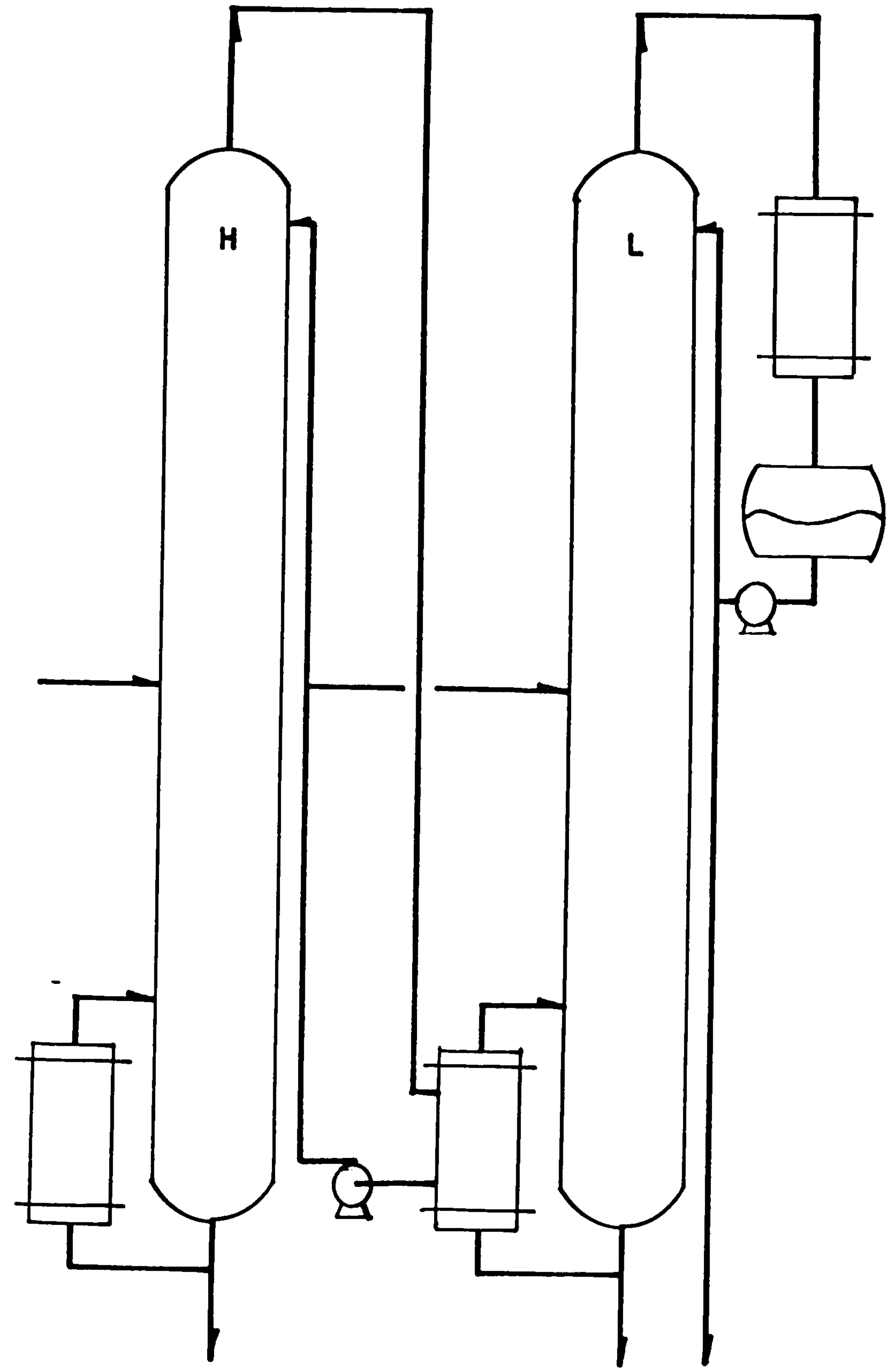

Figure - 4 Configuration 3 (HS/F)

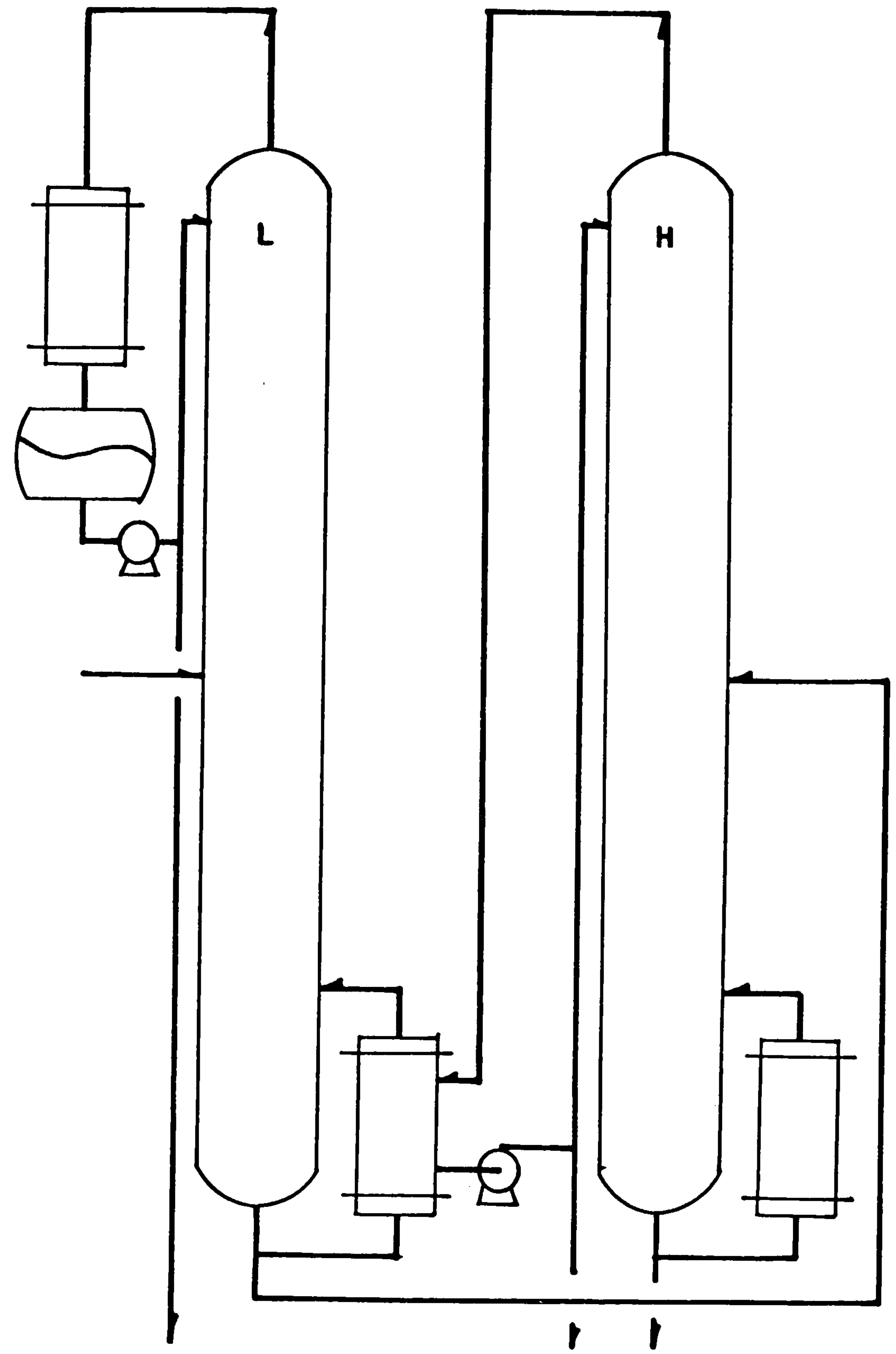

Figure - 5 Configuration 4 (L S / R)

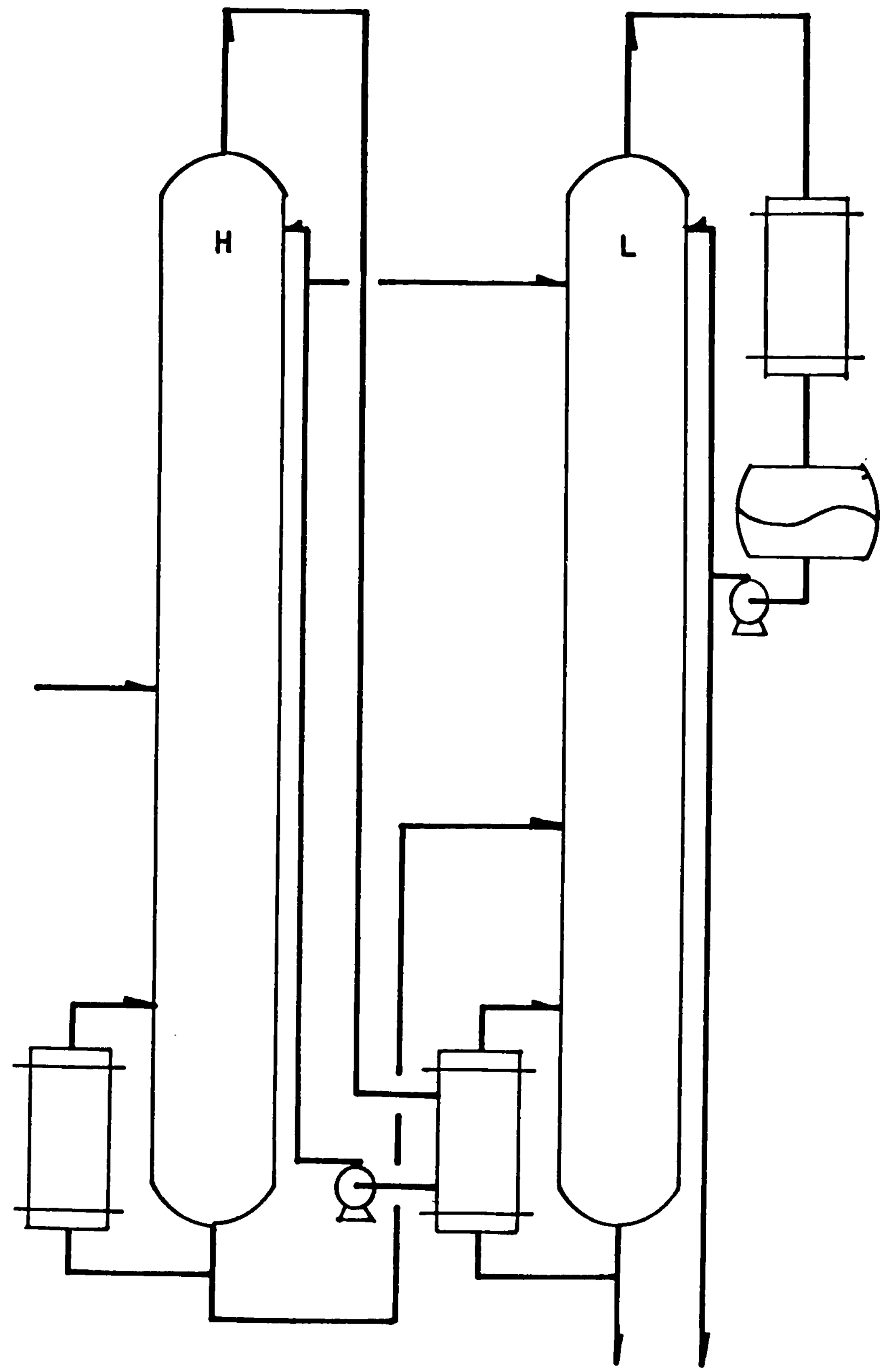

Figure - 6 Configuration 5 (P F/ F)

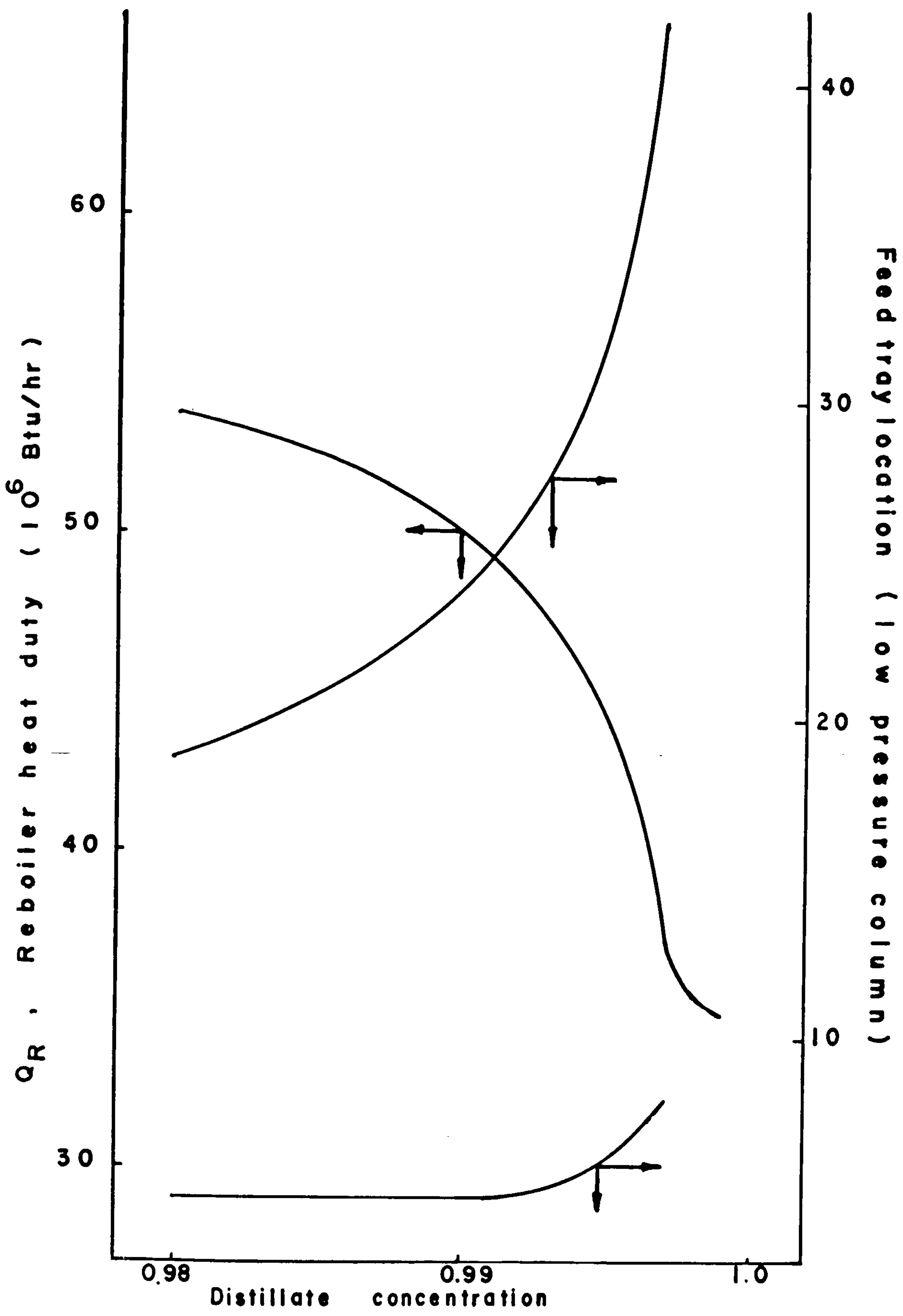

Figure - 7 PF/F goes to LS/F (Z = 0.8)

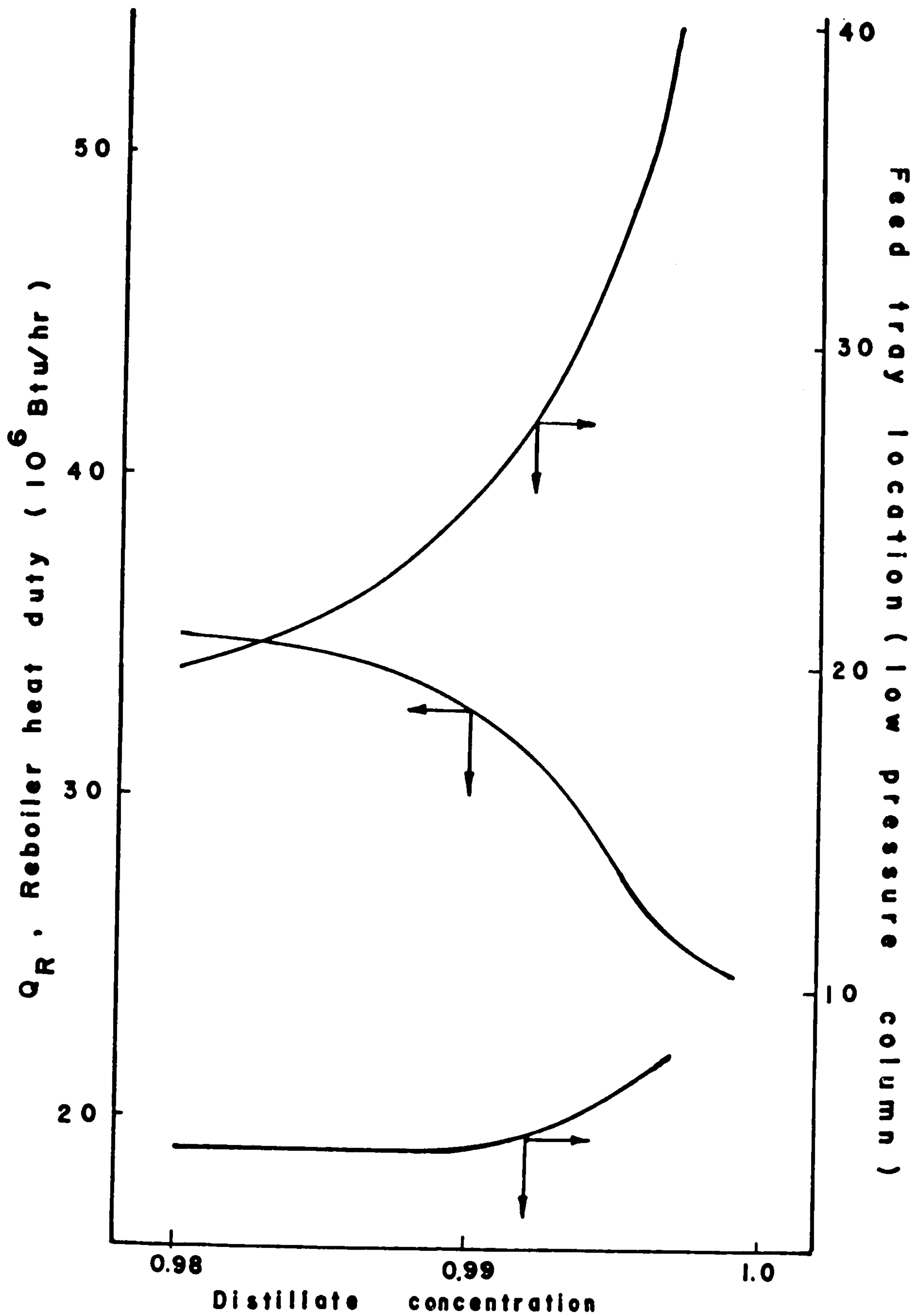

Figure - 8 PF/F goes to LS/F (Z = 0.5)

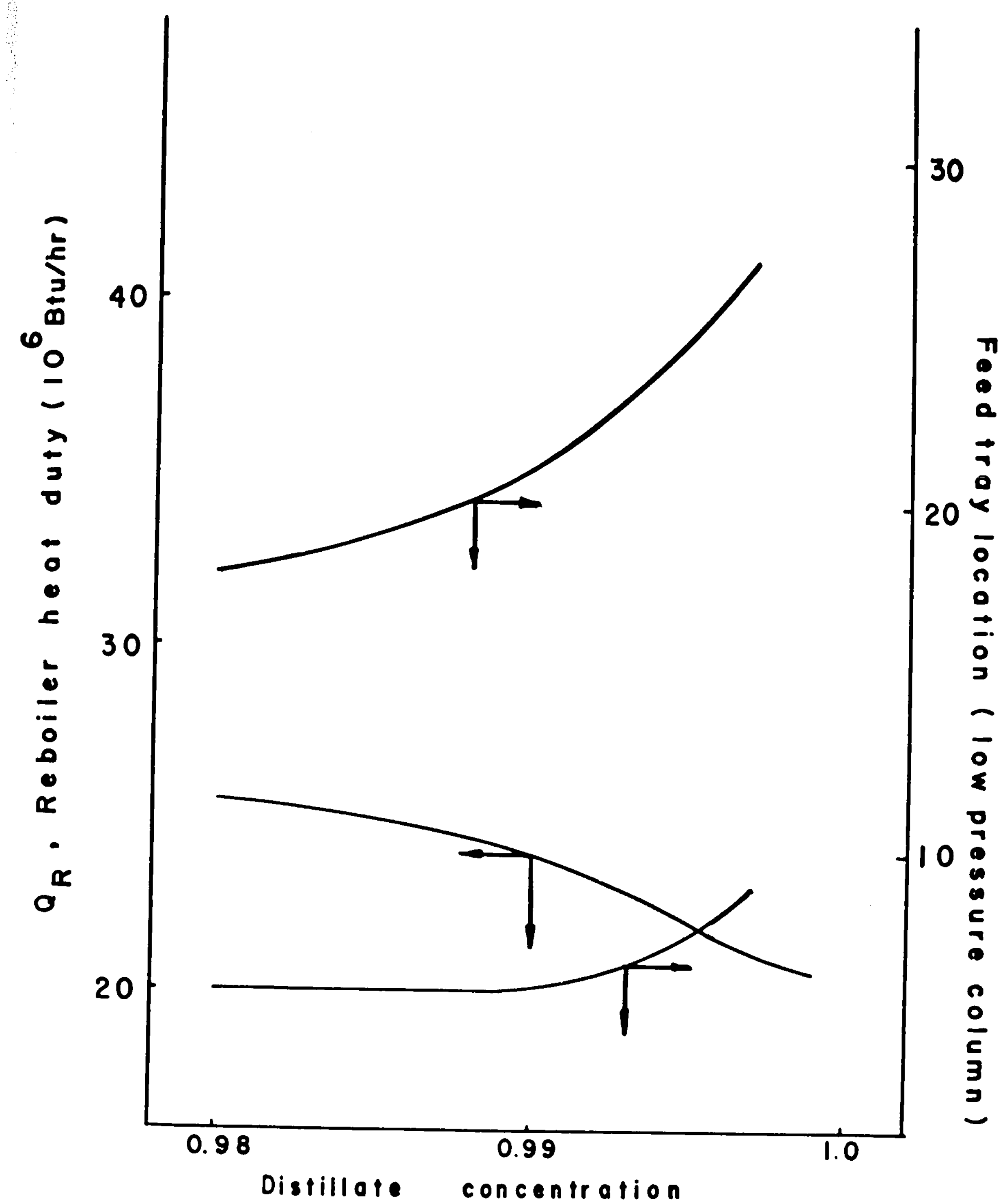

Figure - 9 PF/F goes to LS/F (Z= 0.3)

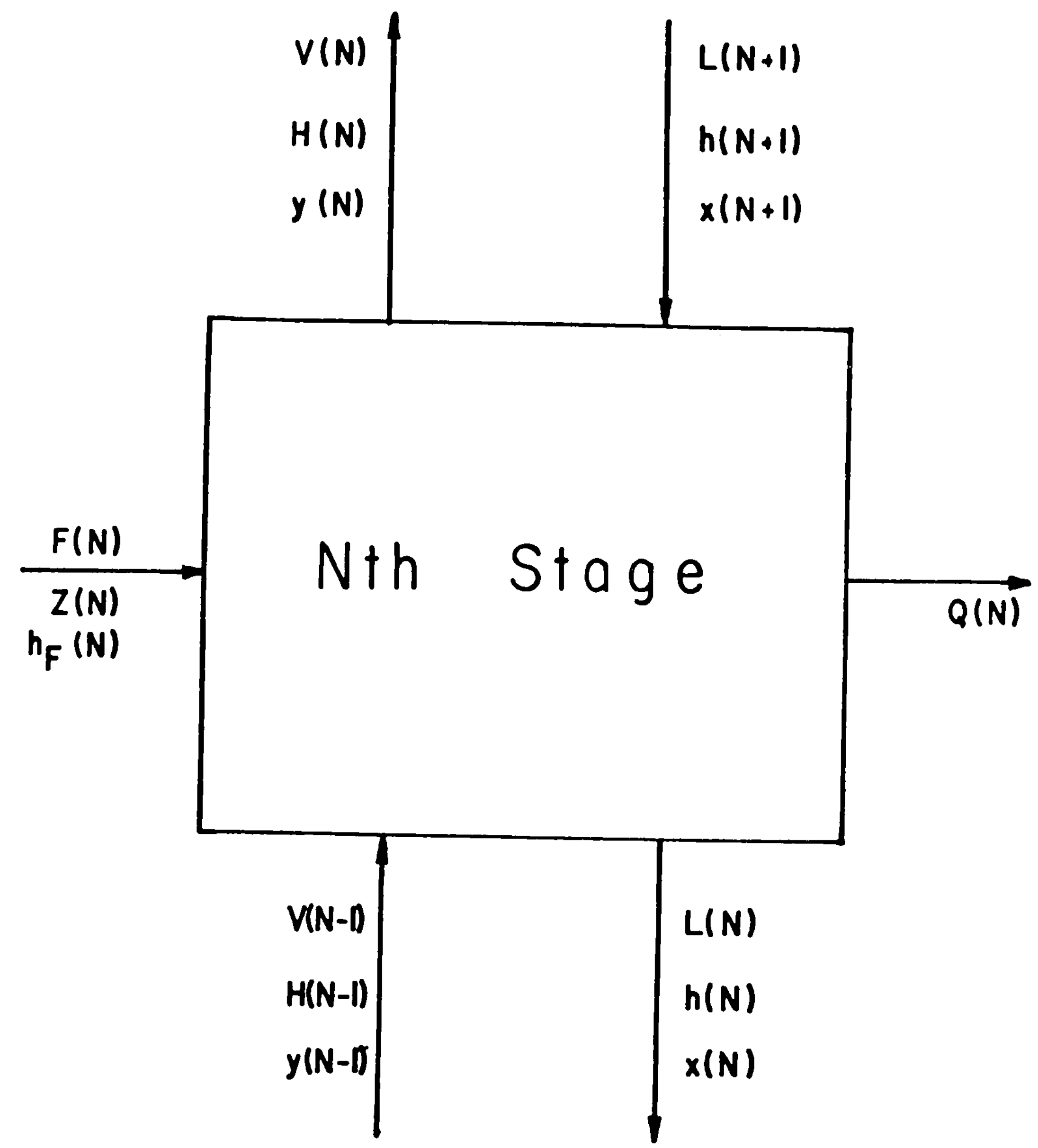

Figure - 10 Ideal equilibrium stage

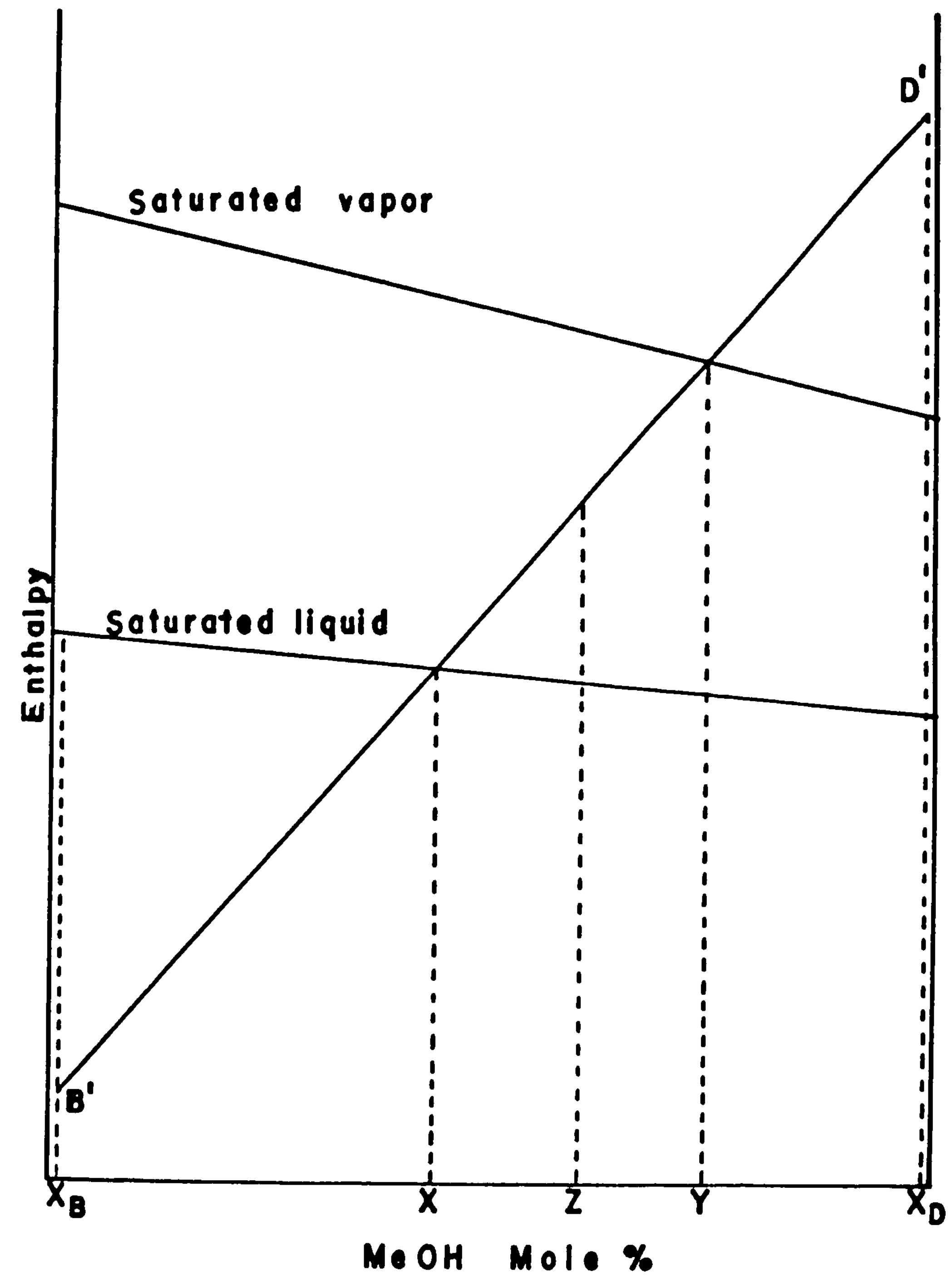

Figure - 11 Enthalpy-concentration diagram

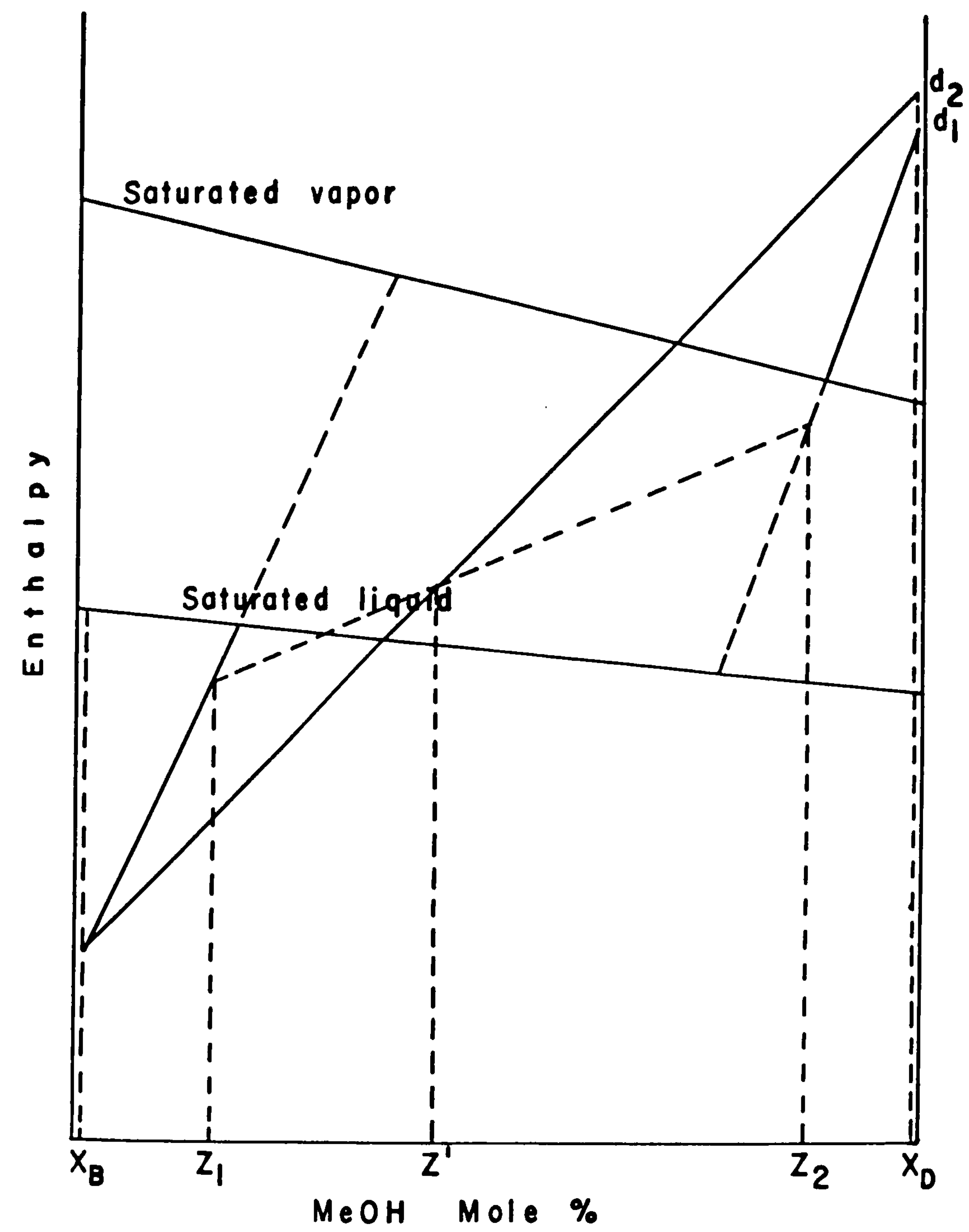

Figure - 12 The Ponchon - Savarit diagram for dual feeds

REFERENCES

1. Freshwater, D. C., "Thermal Economy in Distillation," Trans. Inst. Chem. Eng., 29, 149 (1951)
2. Gmehling, J., and Onken, U., Vapor-Liquid Equalibria Data Collection Aqueous-Organic Systems. Vol.1, Part 1, Dechema, 1977.
3. King, C. J., Separation Process, McGraw-Hill Inc.,1971.
4. Johnson, A. I., and Menzies, M. A., "Modeling and Parametric Studies of a Vapor Recompression Distillation System," Can. J. of Ch.E., 49, 407 (1971)
5. Lenhoff, A. M., and Morari, M., "Design of Resilient Process Plants I. Process Design under Consideration of Dynamic Aspects," Submitted for publication.
6. Mitsuho Hirata, and Shuzo Ohe, and Kunio Nagahana, Computer Aided Data Book of Vapor-Liquid Equalibria, Kodansha and Elsevier Company, 1975.
7. Null, H. R., "Heat Pumps in Distillation," Chem. Eng. Progr., 72, 58, 1976.
8. Rathore, R. N. S., and Van Wormer, K. A., and Powers, G. J., "Synthesis Strategies for Multicomponent Separation with Energy Integration," AICHE, 20, 3, 491,1974.
9. Robertson, J. C., "Energy Conservation in Existing Plants," Chem. Eng., 81, 104, 1974.
10. Robinson, C. S., and Gilliland, E. R., McGraw-Hill Inc. 1950.

11. Scheiman, A. D., "Find Minimum Reflux by Heat Balance," Hydro. Processing, Sep. P.187, 1969.

12. Touloukian, Y. S., and Makita, T., Thermophysical Properties of Matter, Vol.6, Specific Heat, IFI/Plenum, New York, 1970.

13. Tyreus, B. D., "Control of Multi-Effect, Energy conservation Distillation Systems," Ph.D. thesis, Lehigh University, 1975.

14. Van Winkle, M. Distillation, McGraw-Hill, Inc. 1969.

APPENDIX A

A.1. Vapor Pressure Functions

Reference : Mitsuho Hirata et.al. (1975)

Antoine equation :

$$\log_{10} (P_i^o) = A - B / (t + C)$$

$A_{MeOH} = 7.87863$

$B_{MeOH} = 1473.110$

$C_{MeOH} = 230.0$

$A_{H_2O} = 7.96681$

$B_{H_2O} = 1668.210$

$C_{H_2O} = 228.0$

t : temperature, in ^oC

P_i^o : vapor pressure of pure component, in mmHg

A.2. Non-ideality Functions

Reference : Mitsuho Hirata et.al. (1975)

Gmehling J. et.al. (1977)

Activity coefficient is not a strong function of pressure . For low pressure column data at atmospheric pressure were used, for high pressure column data at 5 atm were used.

Van Laar equation :

$$\ln r_1 = A_{12}(A_{21}X_2/(A_{12}X_1+A_{21}X_2))^2$$
$$\ln r_2 = A_{21}(A_{12}X_1/(A_{12}X_1+A_{21}X_2))^2$$

High pressure column :

$A_{12} = 0.7022$

$A_{21} = 0.5830$

Low pressure column :

$A_{12} = 0.8041$

$A_{21} = 0.5619$

(1) Methanol

(2) Water

A.3. Enthalpy Functions

Reference : Touloukian and Makita (1970)

The enthalpy of a mixture is obtained from the pure components by neglecting the mixing effects .

$$H^L_{mix} = xH^L_{MeOH} + (1-x)H^L_{H_2O}$$
$$H^V_{mix} = yH^V_{MeOH} + (1-y)H^V_{H_2O}$$
$$H^L_{MeOH} = -4168.8 - 7.754\ T + 0.02403\ T^2$$
$$H^L_{H_2O} = -1161.0 - 16.43\ T + 0.00126\ T^2$$
$$H^V_{MeOH} = 10048.5 + 5.53\ T + 0.004806\ T^2$$
$$H^V_{H_2O} = 12198.3 + 7.625\ T + 0.00036\ T^2$$

T : tray temperature , in ^{o}F

APPENDIX B

STEADY STATE SIMULATION TECHNIQUES

B.1. Ideal equalibrium stage is shown in Figure 10 . The equations describing the distillation column are as following :

1. Vapor-liquid equalibrium

$$T_1 = f(x_1 , P_1) \qquad (1)$$

2. Physical property (enthalpy)

$$H_1^L = f(x_1 , T_1) \qquad (2)$$

$$H_1^V = f(y_1 , T_1) \qquad (3)$$

3. Total mass balance

Single feed :

$$L_1 = V_B + B \quad \text{(Reboiler)} \qquad (4)$$

$$L_{n+1} = V_n + B \quad \text{(Stripping Section)} \qquad (5)$$

$$L_{n+1} + F = V_n + B \quad \text{(Rectifying Section)} \qquad (6)$$

Dual feeds :

$$L_1 = V_B + B \quad \text{(Reboiler)} \qquad (7)$$

$$L_{n+1} = V_B + B \quad \text{(Stripping Section)} \qquad (8)$$

$$L_{n+1} + F_1 = V_n + B \quad \text{(Middle Section)} \qquad (9)$$

$$L_{n+1} + F_1 + F_2 = V_n + B \quad \text{(Rectifying Section)} \qquad (10)$$

4. Component balance

Single feed :

$$x_1L_1 = V_By_B + Bx_B \quad \text{(Reboiler)} \qquad (11)$$

$$x_{n+1}L_{n+1} = V_ny_B + Bx_B \quad \text{(Stripping Section)} \qquad (12)$$

$$x_{n+1}L_{n+1} + FZ = V_n y_n + Bx_B \quad \text{(Rectifying Section)} \quad (13)$$

Dual feeds :

$$x_1 L_1 = V_B y_B + Bx_B \quad \text{(Reboiler)} \quad (14)$$

$$x_{n+1}L_{n+1} = V_n y_n + Bx_B \quad \text{(Stripping Section)} \quad (15)$$

$$x_{n+1}L_{n+1} + F_1 Z_1 = V_n y_n + Bx_B \quad \text{(Middle Section)} \quad (16)$$

$$x_{n+1}L_{n+1} + F_1 Z_1 + F_2 Z_2 = V_n y_n + Bx_B \quad \text{(Rectifying Section)} \quad (17)$$

5. Energy balance

Single feed :

$$h_1 L_1 + Q_{RBL} = V_B H_B + Bh_B \quad \text{(Reboiler)} \quad (18)$$

$$h_{n+1}L_{n+1} + Q_{RBL} = V_n H_n + Bh_B \quad \text{(Stripping Section)} \quad (19)$$

$$h_{n+1}L_{n+1} + Q_{RBL} + h_F F = V_n H_n + Bh_B \quad \text{(Rectifying Section)} \quad (20)$$

Dual feeds :

$$h_1 L_1 + Q_{RBL} = V_B H_B + Bh_B \quad \text{(Reboiler)} \quad (21)$$

$$h_{n+1}L_{n+1} + Q_{RBL} = V_n H_n + Bh_B \quad \text{(Stripping Section)} \quad (22)$$

$$h_{n+1}L_{n+1} + Q_{RBL} + h_{F_1}F_1 = V_n H_n + Bh_B \quad \text{(Middle Section)} \quad (23)$$

$$h_{n+1}L_{n+1} + Q_{RBL} + h_{F_1}F_1 + h_{F_2}F_2 = V_nH_n + Bh_B$$

(Rectifying Section) (24)

B.2. Algorithm

1. Minimum reflux ratio

Single feed :

Feed with concentration Z was either subcooled , flashed or superheated which might be determined by the position of X, Y, see Figure 11 . Three cases are possible :

(1) $X>Z \longrightarrow$ subcooled (25)

(2) $X<Z<Y \longrightarrow$ flashed (26)

(3) $Y<Z \longrightarrow$ superheated (27)

Figure 11 is shown as case (2) .

Define ,

Slope 1 = (ENTHLP(Y)-ENTHLP(X))/(Y-X) (28)

Slope 2 = (ENTHLP(Z)-ENTHLP(Y))/(Z-Y) (29)

1. Guess a concentration (X)
2. Evaluate Y by bubble point calculation .
3. Check if slope 1 = slope 2 .
4. I. Yes. Then we can calculate minimum reflux ratio , R_{Dm} ;

$$R_{Dm} = (H_D' - H_{y1})/(H_{y1} - h_D) \quad (30)$$

II. No. Change the guess of X and go back to step 2 . Convergence method used here is false-position .

Dual feeds :

Method used was proposed by Scheiman (1969), see Figure 12 .

Assume F_1 controls ;

$$L/V = (y-X_B)/(x-X_B) \quad (31)$$

$$b = (ENTHLP(x)(L/V)-ENTHLP(y))/(L/V-1) \quad (32)$$

$$d_1 = (X_D-X_B)(HIN-b)/(X_F'-X_B)+b \quad (33)$$

Where

$$X_F' = (F_1Z_1+F_2Z_2)/(F_1+F_2) \quad (34)$$

$$HIN = (ENTHLP(F_1)F_1+ENTHLP(F_2)F_2)/(F_1+F_2) \quad (35)$$

Assume F_2 controls ;

$$L/V = (X_D-y)/(X_D-x) \quad (36)$$

$$d_2 = (NETHLP(y)-ENTHLP(x)(L/V))/(1-L/V) \quad (37)$$

Either d_1 or d_2 will be larger . Call the larger value d and use it to calculate the minimum reflux ratio.

$$R_{Dm} = (d-H_{y1})/(H_{y1}-h) \quad (38)$$

2. Reboiler heat duty and condenser heat duty

R_{Dm} multiplied by a number, which is greater than 1, is the operating reflux ratio. Then reboiler heat duty (Q_{RBL}) and condenser heat duty (Q_{CND}) can be calculated .

3. Total tray number and feed tray location

1. Guess a concentration (X_{n+1}) .

2. Evaluate temperature (T) and vapor concentration (Y_{n+1}) through bubble point calculation .

3. Evaluate enthalpy of liquid h(X) and vapor h(Y).

4. Evaluate vapor flow rate through equations (5),(12),(19) for stripping section for single feed, and equations (6),(13),(20) for rectifying section. For dual feeds equations (8), (15),(22) for stripping section , (9),(16),(23) for middle section and (10),(17),(24) for rectifying section.

5. Evaluate liquid flow rate (L_{n+1}) through equation (5) for stripping section for single feed and equation (6) for rectifying section. For dual feed , equation (15) for stripping section, equation (16) for middle section, and equation (17) for rectifying section .

6. Evaluate $(X_{n+1})^{new}$ through equation (12) for stripping section for single feed and (13) for rectifying section . For dual feeds , equation (15) for stripping section, equation (16) for middle section and equation(17) for rectifying section .

7. Check if x is equal to new x .

8. I Yes. Guess $X_{n+2} = X_{n+1} + 0.05$ and go back to step 2 .

II No. Using $(X_{n+1})^{new}$ as new guess and go back to step 2 .

9. Feed tray location is determined when the convergent value of X is greater than the x which was found from minimum reflux ratio calculation .

4. Heat balance between the two columns

After each column has been designed to the specified total tray number , reboiler heat duty of low pressure column will be balanced to the condenser heat duty of high pressure column, considering 5% heat loss, by means of either changing flow rate of distillate or bottoms .